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Flue Gas Desulfurization Pilot Study. Phase I Survey of Major Installations. Appendix 95-I Magnesim Oxide Flue Gas Desulfurization Process

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COMMITTEE ON THE CHALLENGES
OF MODERN SOCIETY

Flue Gas Desulfurization Pilot Study Phase I — Survey of Major Installations - Magnesium Oxide Flue Gas Desulfurization Process

Appendix 95-I

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16. ABSTRACT Magnesium oxide flue gas desulfurization (FGD) is a regenerable process. The sulfur dioxide (SC_2) in the flue gas is absorbed by a magnesium-based slurry, and the resulting magnesium sulfite ($MgSO_3$) is calcined to regenerate magnesium oxide (MgO), which is recycled back to the power plant for reuse in the scrubbing section. The sulfur value from the SO_2-rich by-product gas is recovered.

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PHASE I

Flue Gas Desulfurization Pilot Study Phase I

Survey of Major Installations - Magnesium Oxide Flue Gas Desulfurization Process

APPENDIX 95-I

Lead Country - United States

Prepared by

Frank Princiotta, U.S. Environmental Protection Agency, Washington, D.C.
Richard W. Gerstle, PEDCo Environmental, Cincinnati, Ohio
Edmund Schindler, PEDCo Environmental, Cincinnati, Ohio

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SUMMARY

Magnesium oxide flue gas desulfurization (FGD) is a regenerable process. The sulfur dioxide (SO_2) in the flue gas is absorbed by a magnesium-based slurry, and the resulting magnesium sulfite ($MgSO_3$) is calcined to regenerate magnesium oxide (MgO), which is recycled back to the power plant for reuse in the scrubbing section. The sulfur value from the SO_2-rich by-product gas is recovered.

The chemistry and process design of this system are well developed. Similar processes have been used in the paper pulp industry for many years. One application of this FGD system was first developed in Germany in 1964 and tested in 1968. Full-scale demonstration programs in the United States were recently terminated, but one U.S. unit is still in operation. Two units also are operating in Japan. Initially, many mechanical problems were encountered, as well as design/sizing problems with the solids-handling equipment. Most have been identified and solved.

The magnesium oxide system can best be applied to boilers located in industrial areas where sulfuric acid can be marketed. This can become an additional advantage if

other boilers in the plant or area choose this system because the regeneration facility can serve a number of scrubber systems. The system requires a minimal amount of fresh magnesium oxide and water; thus raw materials are not a limiting factor. Land requirements are relatively small because no sludge is generated. Unless the boiler is very large, the construction of an individual MgO regeneration facility and sulfuric acid plant is generally not feasible.

The advantages and disadvantages of the MgO system are summarized as follows:

Advantages

1. The regeneration of the absorbant (MgO) reduces the cost of raw materials.

2. Disposal problems are relatively minor because the MgO is regenerated and the sulfur is recoverable in a usable form.

3. The sulfur can be recovered as a high-grade acid or possibly elemental sulfur, both of which are marketable.

4. Regeneration can be carried out at a site distant from the power plant, permitting the use of a central regeneration facility to service several FGD units.

5. By maintaining adequate inventories of MgO, the boiler can tolerate extended outages of the regeneration facility without interrupting operation.

6. Modifications and investigations of the first large-scale installations have resulted in improved process reliability.

7. High solubility of $MgSO_4$, high circulation rates, control of slurry composition, high concentrations of crystallization nuclei, and a short residence time reduce scrubber plugging and scaling problems.

Disadvantages/Problems

1. Energy needs are relatively high because of the high temperatures (800° to 1000°C) required in the MgO regeneration step.

2. Reliability at the demonstration plants in the United States has been relatively low since frequent equipment and processing problems have been encountered. Commercial plants in Japan, on the other hand, demonstrated high performance reliability.

3. Experience with the application of the magnesium oxide process to coal-fired boilers is very limited.

4. The centrifuge cake containing $MgSO_3$ can present handling difficulties.

5. Very little data are available on the equilibrium of aqueous magnesium salts.

6. Corrosion/erosion problems in the slurry-handling process of this system are common.

7. Initial and makeup MgO is required in particle sizes of less than 150 μm (100 mesh) to increase reactivity.

8. The scrubber requires liquid-to-gas (L/G) ratios of 5.3 liter/m^3 for absorption and 2.7 liter/m^3 for quenching.

9. Because fly ash reduces the efficiency of the regeneration system, extensive particulate removal is required from a coal-fired boiler.

1.0 BACKGROUND AND INTRODUCTION

This report is one of a series on the status of flue gas desulfurization (FGD) processes in the NATO countries and Japan. The United States, the Federal Republic of Germany, and Norway have acted as lead countries in the gathering of information and for preparation of the reports.

The MgO additive system removes SO_2 from flue gases by scrubbing them with a magnesium-based slurry. A centrifuge removes the solids from the scrubbing solution, after which they are dried and shipped to a regeneration facility for the recovery of their sulfur value. Regenerated magnesium oxide is returned to the power plant for reuse in the scrubbing system.

One of the early MgO processes was developed in 1964 by Grillo-Werke AG in Duisburg-Hamborn, Germany. In 1968 this process was tested in a 7.08 m^3/s, oil-fired boiler at Union Kraft in Wesseling, Germany. The system was also developed and installed by the Mitsui Engineering and Shipbuilding Company in Japan. Although the results were encouraging, for economic reasons the system was later replaced with a lime scrubbing system.

The principal activity in connection with MgO scrubbing in the United States was initiated by a full-scale prototype test facility on oil-fired boiler Unit 6 at the Boston Edison Mystic Station in Everett, Massachusetts. The Chemical Construction Corporation (Chemico) built the scrubbing facility at Everett and the regeneration plant at Rumford, Rhode Island, in April 1972. During the 2-year test program which ended in 1975, operating problems (mostly mechanical) were identified and solved except for corrosion/erosion problems. Both plants indicated the potential for achieving 90 percent availability with over 90 percent SO_2 removal efficiency.

The MgO process is quite similar to the magnesium-based pulping system, generally called the "Magnefite" process. The main differences between the Magnefite and the MgO FGD system are that (1) the pulp mill gas contains more SO_2 (1% or more vs. 0.1 to 0.4%), (2) the object in pulping is to produce soluble $Mg(HSO_3)_2$, whereas in stack gas scrubbing $MgSO_3$ is desired, and (3) the calcining step in the pulping operation involves decomposition of complex metal organic compounds rather than simple $MgSO_3$.

Two MgO scrubbing plants are currently operating in Japan, and one demonstration plant with 120 MW of generating capacity is operating in the United States.

Another 2-year test program at Dickerson Generating Station of the Potomac Electric and Power Co. operated from September 1973 to August 1975 on a coal-fired utility boiler. This FGD system treated half of the flue gas from the 190-MW boiler while burning 2 percent sulfur content coal. The magnesium sulfite crystal from the coal-fired boiler was different from that obtained at the oil-fired Mystic Station and required a change in calciner operation.

2.0 GENERAL PROCESS DESCRIPTION

The principles of all MgO scrubbing systems are essentially the same. A magnesium-based slurry is contacted with flue gas containing SO_2. The SO_2 is absorbed and forms an insoluble product. After this product is removed from the scrubbing solution by a centrifuge, it is dried and shipped to a separate regeneration facility for the recovery of its sulfur value and the MgO. The regenerated MgO is recycled back to the power plant for reuse in the scrubbing system. Figure 1 presents an example flow diagram and material balance of a 500-MW coal-fired boiler with a MgO desulfurization plant and a sulfuric acid by-product plant.*

2.1 PROCESS CHEMISTRY

The overall reactions that take place in the absorber are

$$MgO + SO_2 + 6\ H_2O \rightarrow MgSO_3 \cdot 6H_2O \quad (1)$$

$$MgO + SO_2 + 3\ H_2O \rightarrow MgSO_3 \cdot 3H_2O \quad (2)$$

Some $MgSO_4$ is also formed in the absorber as a result of partial oxidation of the $MgSO_3$ by the reaction

$$MgSO_3 + 1/2\ O_2 + 7\ H_2O \rightarrow MgSO_4 \cdot 7H_2O \quad (3)$$

* Other methods of final sulfur recovery can also be used.

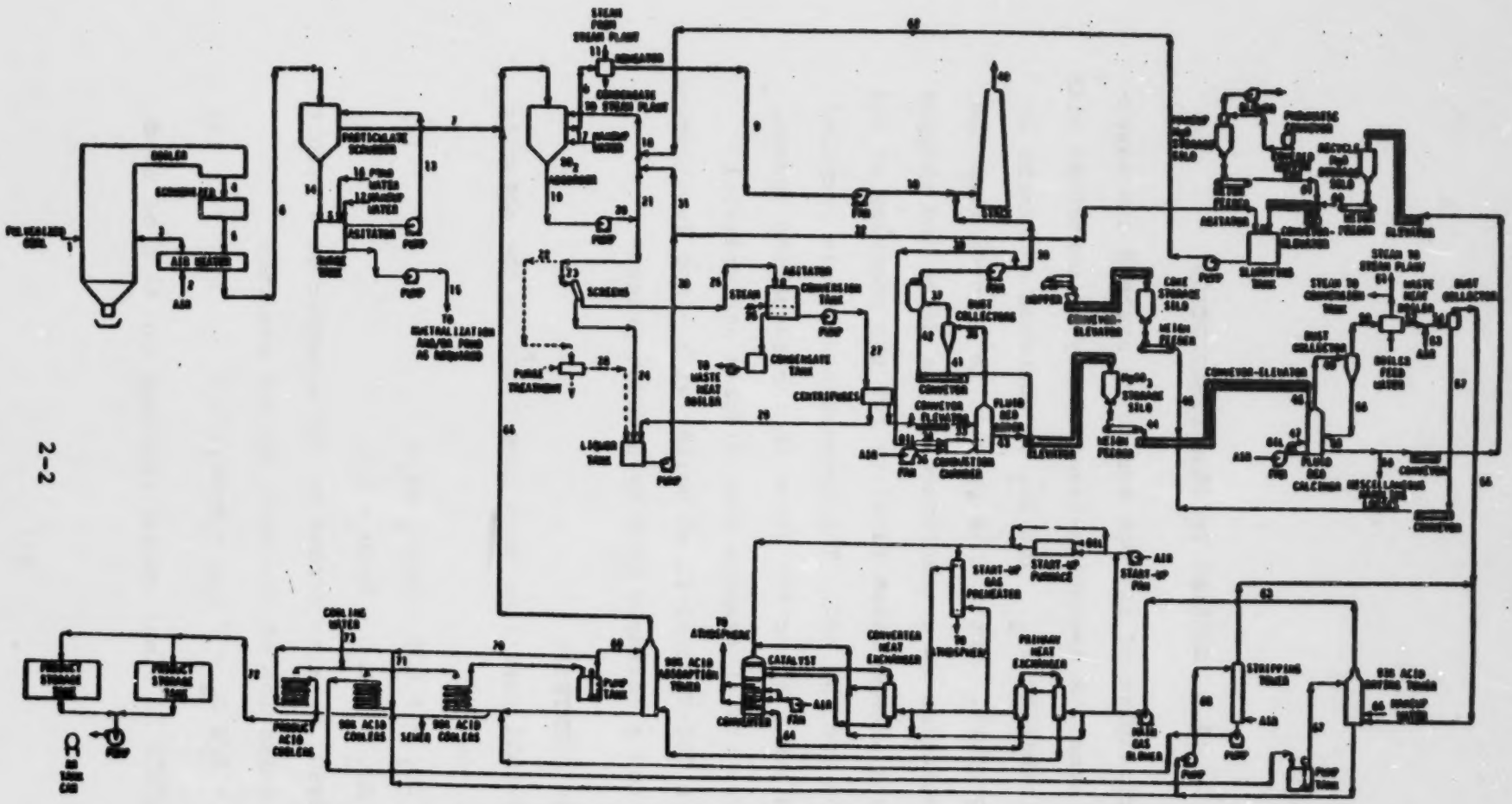

Figure 1. Example Magnesia slurry - regeneration process including H_2SO_4 plant. Flow sheet and mass balance.

(Reference: Detailed Cost Estimates for Advanced Effluent Desulfurization Processes by Tennessee Valley Authority.)

Stream No. / Description	1 Coal to boiler	2 Combustion air to air heater	3 Combustion air to boiler	4 Gas to econo-mizer	5 Gas to air heater	6 Gas to particulate scrubber	7 Gas to SO_2 absorber	8 Gas to reheater	9 Gas to fan	10 Gas to stack
Rate, 1000 kg/h	170	2049	1848	1998	1998	549.8	574.7	584.7	584.7	584.7
Nm^3/s		464	419	445	445	123	132	135	135	135
m^3/s, Liquid										
Particulates, kg/h				15,300	15,300	3826	19.5	25.8	25.8	25.8
Temperature, °C		433	279	477	374	154	52.8	52.8	71.1	70.9
Specific gravity										
Viscosity, pascal-seconds										
Undissolved solids, %										
pH										
Sulfur as SO_2, kg/h				11,335	11,335	2834	2645	284	1134	1134

Stream No. / Description	11 Steam to gas reheater	12 Makeup water to part. scrbr.	13 Recycle slurry to part. scrbr.	14 Part. scrbr. slurry to surge tank	15 Particulate slurry product	16 Recycle pond water	17 Makeup water to SO_2 absorber	18 Recycle slurry to SO_2 absorber
Rate, 1000 kg/h	7.12	25	1107	1086	25.4	21.6	3.190	1562
Nm^3/s								
m^3/s, Liquid		0.00694	0.2991	0.2933	0.0065	0.006	0.0009	0.3987
Particulates, kg/h								
Temperature, °C	243							
Specific gravity			1.03	52.8	1.09			
Viscosity, pascal-seconds								
Undissolved solids, %			5		15			
pH				3				
Sulfur as SO_2, kg/h		8270	189					

Figure 1 (continued).

(See notes at end of figure.)

Stream No. / Description	19 SO_2 absorber slurry to pump	20 Recycle and product slurry	21 Recycle slurry	22 Slurry to purge treatment	23 Slurry to screen	24 Underflow to liquor tank	25 Overflow to conservation tank	26 Steam to conservation tank	27 Slurry to centrifuge
Rate, 1000 kg/h	1568	1568	1444	Depends upon magnesia and makeup water impurities fly ash entrainment, etc.	123	100	23	5.81	46.3
Mg^3/s									
m^3/s, Liquid	4000	.4000	.3686		0.032	0.027	0.00514		0.0107
Particulates, kg/h									
Temperature, °C	52.8							186	82.2
Specific gravity		1.09				1.05	1.25		1.20
Viscosity, pascal-seconds									
Undissolved solids, %		10				3.1	40		31
pH		8							
Sulfur as SO_2, kg/h	2561	2561							

Stream No. / Description	28 Treated liquor to liquor tank	29 Liquor to liquor tank	30 Liquor to recycle	31 Recycle liquor	32 Liquor to slurrying tank	33 Feed to dryer	34 Oil to combustion chamber	35 Air to combustion chamber	36 Dryer gas to bag cyclone
Rate, 1000 kg/h	see Stream 22 note	31.1	463.1	101	58.5	30.1	1.335	27.4	61.7
Mg^3/s								6.23	16.52
m^3/s, Liquid		0.0082	0.1226	0.0268	0.0155				
Particulates, kg/h									2703
Temperature, °C									204
Specific gravity		1.06	1.05						
Viscosity, pascal-seconds									
Undissolved solids, %		4.6	3.3						
pH									
Sulfur as SO_2, kg/h									

Figure 1 (continued).

Stream No.	37	38	39	40	41	42	43	44
Description	Dryer gas to bag collector	Recycle gas to combustion chamber	Dryer gas to stack	Combined gases to atmosphere	Cyclone dust to conveyor	Bag collector dust to conveyor	Dryer product to conveyor	Feeder discharge to conveyor
Rate, 1000 kg/h	61.7	20.7	40.9	2380	2.298	0.408	15.2	17.9
Mg^3/s	16.52	5.57	10.95	551				
m^3/s, Liquid								
Particulates, kg/h	405	0.64	1.3	104				
Temperature, °C				79.4			204	
Specific gravity								
Viscosity, pascal-seconds								
Undissolved solids, %								
pH								
Sulfur as SO_2, kg/h							10,200	10,200

Stream No.	45	46	47	48	49	50	51	52	53
Description	Coke to conveyor	Feed to calciner	Oil to calciner	Air to calciner	Calciner gas to cyclone	Calciner gas to waste heat boiler	Steam to steam plant	Calciner gas to quench air	Quench air to calciner gas
Rate, 1000 kg/h	0.099	18.1	1.461	21	32.5	32.5	1.465	32.5	28.4
Mg^3/s				4.77	6.18	6.18		6.18	6.51
m^3/s, Liquid									
Particulates, kg/h					898.1	135		135	
Temperature, °C					871		186	369	
Specific gravity									
Viscosity, pascal-seconds									
Undissolved solids, %									
pH									
Sulfur as SO_2, kg/h		10,200			10,200	10,200		10,200	

Figure 1 (continued).

Stream No. / Description	54 Combined gases to bag collector	55 Combined gases to H_2SO_4 unit	56 Cyclone dust to calciner	57 Bag collector dust to conveyor	58 Calciner product to conveyor	59 Misc. handling losses	60 Recycle MgO to conveyor	61 Makeup MgO to conveyor	62 MgO slurry to SO_2 absorber
Rate, 1000 kg/h	60.8	60.8	0.7634	0.133	7.67	0.109	7.53	0.136	16.6
Mg^3/s	12.70	12.70							
m^3/s, Liquid									0.00414
Particulate, kg/h	135	1.36							
Temperature, °C	204				107				
Specific gravity									1.11
Viscosity, pascal-seconds									
Undissolved solids, %									15
pH									
Sulfur as SO_2, kg/h	10,200		0						0

Stream No. / Description	63 Drying tower outlet gas	64 Converter outlet gas	65 Absorption tower outlet gas	66 Makeup water to drying tower	67 Recycle acid to drying tower	68 Recycle acid to stripping tower	69 Recycle acid to absorption tower	70 Recycle acid to coolers
Rate, 1000 kg/h	58.5	58.5	46.7	0.519	455.9	58.5	450	70.3
Mg^3/s	11.85	11.37	10.38					
m^3/s, Liquid				0.000144	0.07003	0.009	0.07003	0.011
Particulate, kg/h								
Temperature, °C	37.8	449	71.1		37.8	72.8	71.1	
Specific gravity					1.81	1.77	1.79	
Viscosity, pascal-seconds								
Undissolved solids, %								
pH								
Sulfur as SO_2, kg/h			800					

Figure 1 (continued).

Stream No.	71	72	73
Description	Recycle acid to 93% acid coolers	Product acid to storage	Cooling water to acid coolers
Rate, 1000 kg/h	55.8	14.6	1079
Mg^3/s			
m^3/s, Liquid	0.009	0.002	0.3001
Particulates, kg/h		1.36	
Temperature, °C		37.8	
Specific gravity		1.82	
Viscosity, pascal-seconds			
Undissolved solids, %			
pH			
Sulfur as SO_2, kg/h		9400	

Notes:

1. Calculations based on:
   a. 105% stoichiometric magnesia
   b. 3.5% sulfur in coal (dry basis)
   c. 12% ash coal (as-fired basis)
   d. 92% of sulfur in coal evolves as SO_2
   e. 75% of ash in coal evolves as fly ash
   f. 99.5% removal of particulates to scrubber
   g. 90% SO_2 removal
2. Particulates should be added to gas to get total stream rate.
3. Stream numbers 6-21, 23-25, 3, and 62 are one of four similar streams.
4. Stream numbers 27 and 29 are one of two similar streams.

Figure 1 (continued).

The SO_2 is also absorbed by $MgSO_3$ according to the reactions

$$MgSO_3 + H_2O + SO_2 \rightarrow Mg(HSO_3)_2 \quad (4)$$

and

$$Mg(HSO_3)_2 + Mg(OH)_2 + 10\ H_2O \rightarrow 2MgSO_3 \cdot 6H_2O \quad (5)$$

Reactions (4) and (5) are favored in an acidic slurry, with a pH of about 6.

After the crystals are dried, they are calcined to generate SO_2 gas and regenerate MgO. Coke is added to reduce $MgSO_4$ to MgO and SO_2. The overall calcining reactions are

$$MgSO_3 \rightarrow MgO + SO_2 \quad (6)$$

$$MgSO_4 + 1/2C \rightarrow MgO + SO_2 + 1/2CO_2 \quad (7)$$

The calcining temperature is generally between 800° and 1000°C.

According to the molecular weights of MgO and SO_2, the theoretical requirement is one mole of MgO per mole of SO_2 removed. Assuming a 1.1 excess stoichiometric requirement and 97 percent purity of MgO, the actual MgO required is

$$\frac{40.3}{64} \times \frac{1.1}{0.97} = 0.71 \text{ kg MgO/kg of } SO_2.$$

Regenerated MgO is recycled back to the scrubbing system.

2.2 DESCRIPTION OF EQUIPMENT COMPONENTS

The MgO FGD process consists of two systems: the SO_2 scrubbing system and the MgO regenerating system.

2.2.1 Process Particulate Removal

The flue gas enters a venturi scrubber designed to cool the gas to 53°C and to remove particulates from the gas stream. The unit is designed to circulate water, some of which is purged from the recirculation pump discharge to the wastewater treatment system. If an electrostatic precipitator (ESP) is used, a flue gas cooler and a humidifier are required to condition the gas stream prior to absorption. Venturi particulate removal systems require additional power because of the higher pressure drop in the unit.

2.2.2 SO_2 Scrubber

The SO_2 scrubber is the primary SO_2 removal unit. Designs include the turbulent contact absorber (TCA) (Onahama, Japan) and venturi-type scrubbers (Chemico and United Engineers). Figure 2 presents a schematic drawing of a single-stage venturi scrubber combined with an SO_2 absorber.

The scrubber must be constructed of materials resistant to corrosion, erosion, and scaling. Most scrubber bodies are mild carbon steel and have an acid-resistant coating such as fiberglass-reinforced polyester (FRP), polyurethane, rubber, or glass flake. Scrubber internals may be stainless steel, high-nickel alloys, FRP, etc. No one material seems to be superior to the others.

Scrubber module size and the number of modules per

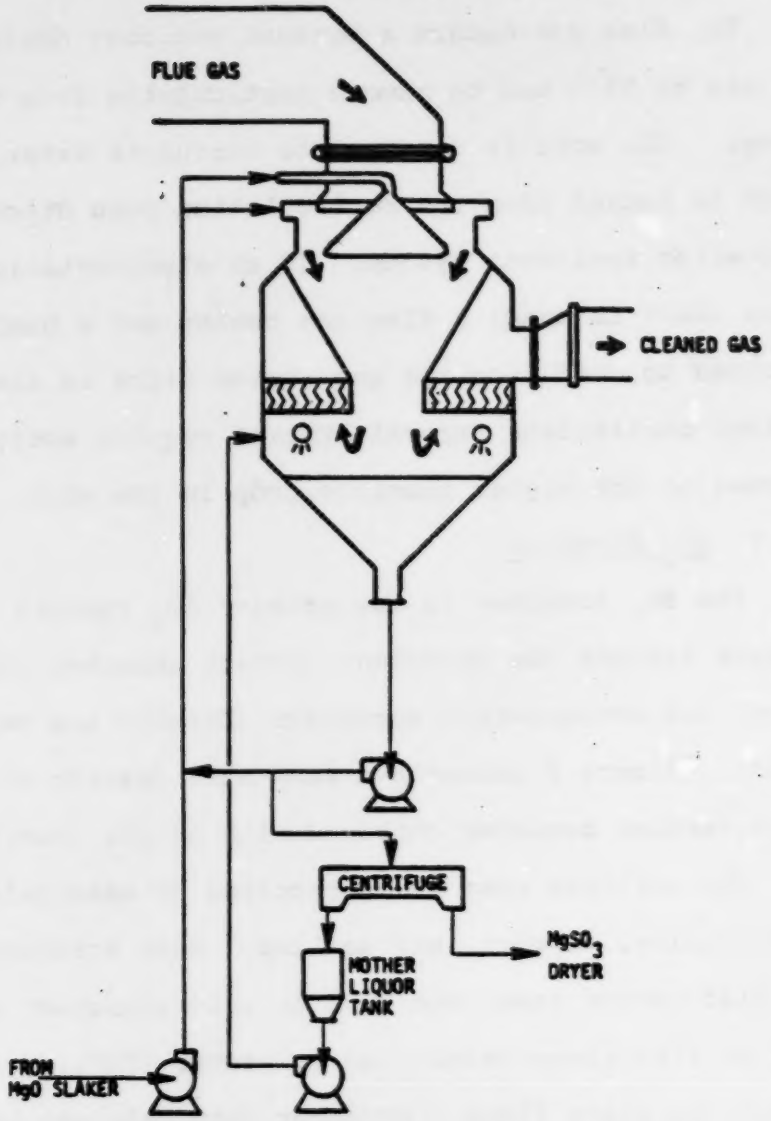

Figure 2. Combination venturi scrubber and absorber.

system have a direct relationship to turndown* requirements, system availability, and gas-liquid distribution. As boiler loads fluctuate, the scrubbing rate must change to maintain optimum performance. One method of accomplishing turndown is to shut down some of the scrubber modules as the load decreases--the more modules in the system, the smoother the transition. Scrubber modules not being used can be scheduled for cleaning and maintenance, thereby reducing overall scrubber downtime. Multiple scrubber modules also have the advantage of using smaller scrubber module cross-sections, which promotes uniform gas-liquid distribution and improves efficiency. Scrubber module sizes range from 25 to 165 MW.

2.2.3 Demister

A demister is necessary to remove entrained water from the scrubber outlet gas, to reduce downstream equipment corrosion and reheater energy requirements. The water droplets are generally large enough to be removed with simple changes in flue gas direction, and the chevron or "Z" type mist eliminator is widely used. Demisters are constructed of stainless steel, fiberglass-reinforced polyester, or other plastic material and can be installed in the horizontal or vertical position.

---

* Reduction in boiler output.

2.2.4 Reheater

The stack gas emerging from the SO_2 scrubber has a temperature of about 54°C and is saturated with water vapor. Reheating of the gas is generally necessary to avoid water condensation and corrosion and to improve plume dispersion. Reheating can be accomplished by 1) installing a gas or low-sulfur oil burner that exhausts directly into the stack, 2) installing steam coils to heat flue gas, 3) using steam coils to heat air and injecting it into the stack, or 4) by-passing untreated hot flue gas.

2.2.5 $MgSO_3$ Recovery Equipment

The recovery of the $MgSO_3 \cdot 6H_2O$ formed in the scrubber surge tank is begun by diverting a side stream from the main scrubber circulation loop to a thickener. The thickener is not an essential element in the system, but it seems to improve the centrifuge operation by providing constant feed concentration.

A wet cake of $MgSO_3 \cdot 6H_2O$, $Mg\ SO_3 \cdot 3H_2O$, and $MgSO_3 \cdot 7H_2O$ crystals is recovered in a stainless steel, solid-bowl centrifuge. These wet crystals are discharged from the centrifuge through a vertical chute into a screw feeder, then into a rotary-kiln. Anhydrous $MgSO_3$ and some $MgSO_4$ are discharged from the dryer and conveyed to storage silos. Dryer temperatures range from 176° to 232°C.

2.2.6 MgO Regenerating Plant

The anhydrous $MgSO_3$ and some $MgSO_4$ are heated in a calciner to regenerate MgO. A rotary-kiln or fluidized-bed calciner can be used for this purpose. A fluidized-bed reactor generally provides a more uniform bed temperature and more surface area for reaction. An experimental fluidized-bed calciner for MgO regeneration has shown favorable results and may be used commercially in the near future.

The MgO is separated from the gas stream by cyclones, and the stack gas is cleaned by a combination of a venturi scrubber and a packed-bed scrubbing tower before it is admitted to the sulfuric acid plant. The regenerated MgO is reused in the scrubbing system.

2.3 OPERATING PARAMETERS

2.3.1 Chemical Parameters

The crystals of hexahydrate formed by reaction (1) are usually large (200 μm), and the material is damp and sand-like. The crystals of trihydrate are much smaller and have a mud-like consistency; therefore, they are centrifuged less efficiently and result in a wetter cake which agglomerates in the dryer.

The nature of the hydrate formation depends somewhat on the temperature in the absorber. The hexahydrates are stable at room temperature. As the temperature increases,

however, the hexahydrates gradually change to more stable trihydrates. A recent laboratory study shows that 1) the transition temperature is from 60° to 63°C (higher than previously known), 2) the hexahydrates persist even above the transition temperature, and 3) the rate of the transition is very slow.

Magnesium sulfite is formed in the scrubber by the reaction between the SO_2 and the alkaline slurry; this product, however, is not very soluble in water. When the scrubbing slurry is acidic, very soluble $Mg(HSO_3)_2$ is formed. Thus, the fundamental chemical reaction in the scrubber tower results in the conversion of a slightly soluble material to a very soluble material. It is quite possible that this approach could eliminate plugging problems in the scrubber. The pH of the scrubbing slurry is controlled by regulating the rate of addition of $Mg(OH)_2$ slurry.

The equilibrium vapor pressure of SO_2 over a slurry of $MgSO_3$ in a slightly acidic, dilute solution of $Mg(HSO_3)_2$ is higher than that over a slurry of alkaline MgO or $Mg(OH)_2$ in water. It is possible, therefore, that the removal of SO_2 from stack gas may not be as effective with an acidic slurry as with an alkaline slurry.

In practice, however, the scrubbing slurry does not approach the point of saturation or equilibrium, and the SO_2 removal capability of the magnesium oxide scrubbing system is not limited. A recent laboratory study demonstrated no great difference between SO_2 removal at pH's of 6 and 8.

2.3.2 Physical Parameters

An MgO FGD system can achieve SO_2 collection efficiencies of 90 percent on coal-fired boilers. Key operating parameters include the following:

- Venturi scrubbers operating at pressure drops of 2500 pascals or greater with an L/G ratio of 5.4 to 6.8 liter/m^3 have achieved collection efficiencies in excess of 90 percent. A TCA scrubber applied on smelter reverberatory furnace off-gas also achieved collection efficiencies of 99 percent and greater.
- Slurry pH (measured in the absorber discharge tank) should be maintained in the range of 6.0 to 7.5.
- The MgO should be slaked (dissolved) in water in a heated agitation tank. Sometimes it requires preslaking and dilution before it is added as makeup to the absorber.
- Prescrubbing of the flue gases with water for residual fly ash removal and cooling will minimize the formation of the trihydrate of $MgSO_3$ and insure better handling, drying, and calcining of the product.

The flue gas is reheated from 54°C to between 71° and 98°C (maximum 110°C) to prevent condensation and enhance dispersion. Reheating requires approximately 1100 J/°C of heat input per kg of flue gas.

3.0 CURRENT STATUS, ASSESSMENT, AND PLANS

Two magnesium oxide FGD systems are operating in Japan. One treats smelter gas and the second treats gases from oil-fired boiler furnaces and Claus furnaces. In the United States, the operation of two demonstration plants was terminated after completion of 2-year tests during which SO_2 removal efficiencies of more than 90 percent were attained. A third system has been in intermittent operation since 1975. This system at the Philadelphia Electric Company's Eddystone Station has been subject to numerous problems. Many of these have been mechanical such as pump seal leaks, plugged valves, and centrifuge erosion. The SO_2 removal efficiency has, however, been high and based on modifications made to the existing system; the owner plans to install three more similar systems.

The magnesium oxide FGD process is a viable system and the chemistry is well developed. The technology has been adequately demonstrated for oil-fired boilers, but insufficient experience is available to ascertain its potential usefulness on coal-fired systems. The system has no significant solid waste and wastewater problems and therefore

can be located in populous areas where land costs are high. The process is especially suited to industrial areas with a market for H_2SO_4. The system is also adaptable to multiple boilers, small or large, each using only one centrally located regeneration facility.

The system requires no raw materials except water and a small amount of MgO to replenish attrition of the initial charge. Although the system still has some mechanical problems, availability of one Japanese plant has been as high as 95 percent; and SO_2 removal efficiencies are from 99.5 to over 99.8 percent from a concentrated (2 to 3% SO_2) smelter gas stream.

The major disadvantages of this system are the relatively high energy requirements and insufficient experience on coal-fired boilers, where fly ash could lead to problems.

One U.S. designer of MgO FGD systems guarantees SO_2 and particulate removal efficiencies and opacity compliance on oil- and coal-fired facilities for a period of 12 to 18 months. This company also offers financing, operating, and maintenance services.

Table 1 summarizes the status of surveyed MgO systems. Collection efficiency and availability ranges on the table refer only to the surveyed systems, but the rest of the

Table 1. SUMMARY OF STATUS OF SURVEYED MAGNESIUM OXIDE FGD SYSTEMS

All Magnesium Oxide FGD Systems	
Number of systems in operation[a] December 1977	4
Total number of systems built	6
Total number of systems surveyed	4
Year first placed in commercial service	1971
Number of utility systems using process	1
Total MW of utility systems	120
Number of industrial combustion systems	1
Surveyed Systems	
Average SO_2 removal efficiencies	Over 90%
Range of fly ash removal efficiency for FGD	54-77%
Range of average availability factors[b]	30-100%

[a] Two noncombustion sources.

[b] See Appendix B for availability factor estimation procedure.

information projects the status of all MgO systems through December 1977. Table 2 presents significant characteristics of each surveyed facility. Questionnaires in the Appendix contain more detailed information.

Table 2. STATUS OF SURVEYED FGD INSTALLATIONS USING MAGNESIUM OXIDE PROCESS

Name of plant and company	Location	Size MW	Fuel type	Characteristics		Flue gas treated from boiler,		Year FGD system started	SO_2 removal efficiency, %	Disposition of end products	Energy required, % total input	Availability, %
				S, %	Heat value, 10^6 J/kg	m^3/s	% of total					
Mystic Station Boston Edison	Everett, Massachusetts	155	No. 6 fuel oil	2.5	43.2	178	100	1972	91.5	$MgSO_3$	2.1[a]	50-76
Dickerson Potomac E.P.	Dickerson, Maryland	95	Coal	2.0	27.2	93	100	1973	90.7	$MgSO_3$	1.6[b]	N.A.
Eddystone Philadelphia E.	Eddystone, Pennsylvania	120	Coal		28	91	33	1975	over 90	$MgSO_3$	1.6[c]	30
Chiba Refinery Idemitsu Kosan Co.	Chiba, Japan	160	Heavy oil C (2 boilers)	3.8	43.1	128[d]	47[e]	1974	over 95	Sulfur	5.1[f]	100
			Claus furnace gas									

[a] Calciner not included, reheat data not available.

[b] Energy required is estimated; calciner not included, reheat data not available.

[c] In addition 3.4% reheat energy required, energy required is estimated; calciner not included.

[d] Combined gas has 0.285% SO_2.

[e] 100% of the gas from 2 boilers.

[f] Includes calciner.

4.0 RECENT TECHNOLOGICAL DEVELOPMENTS AND SOLUTIONS TO PROBLEM AREAS

4.1 RECENT DEVELOPMENTS

The design of scrubbers and calciners for MgO FGD equipment has noticeably improved.

In the United States, the first MgO process on a boiler used a single-stage venturi scrubber. No particulate removal equipment was used in this installation, however, because it burned fuel oil. The second MgO installation, the first coal-fired application, involved a two-stage venturi device (Figure 3). In this design, particulates were removed in the first stage in a variable-throat venturi and SO_2 was absorbed by magnesium-based slurry in a fixed-throat venturi in the second stage. Particulates were removed in the variable-throat venturi using water and operating at 121°C. The gas was saturated and cooled to 49°C as the fly ash was removed.

In a third installation, a rectangular venturi-rod scrubber is used (Figure 4). This design essentially has two scrubbing stages in one scrubber. In each stage, a slurry of magnesium sulfite is sprayed upward through openings

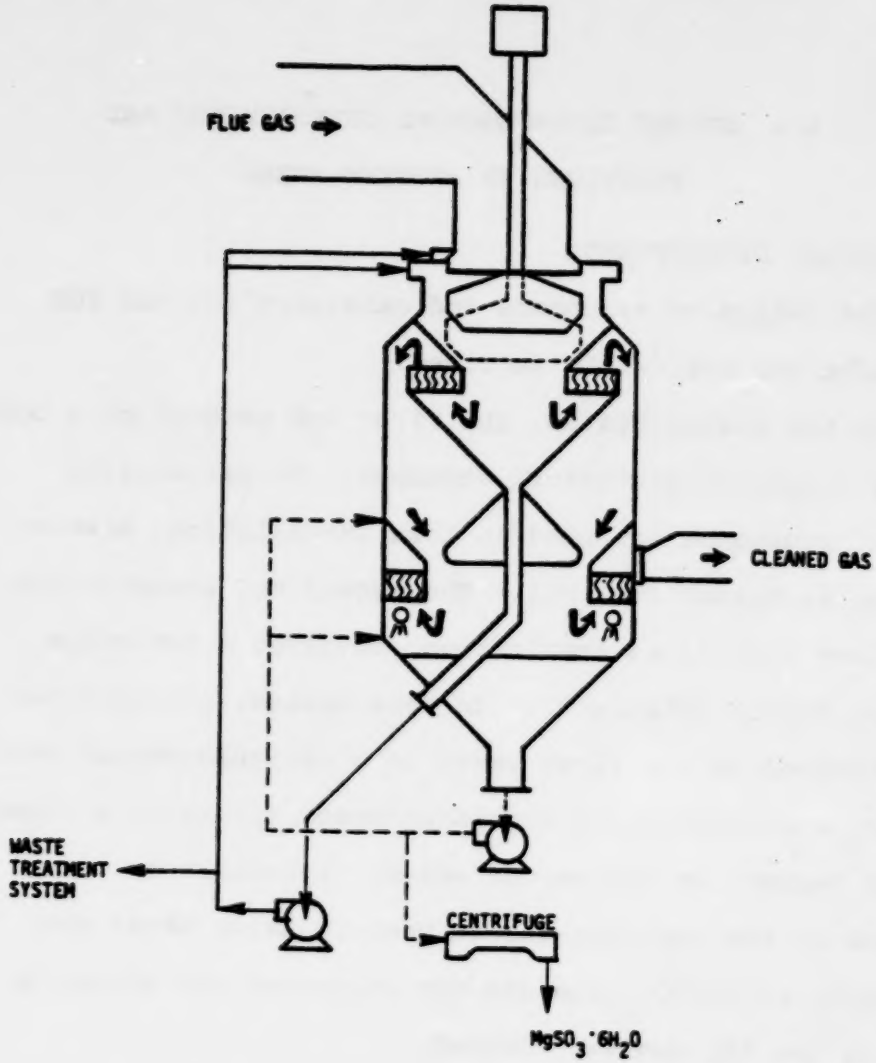

Figure 3. Two-stage venturi scrubber.

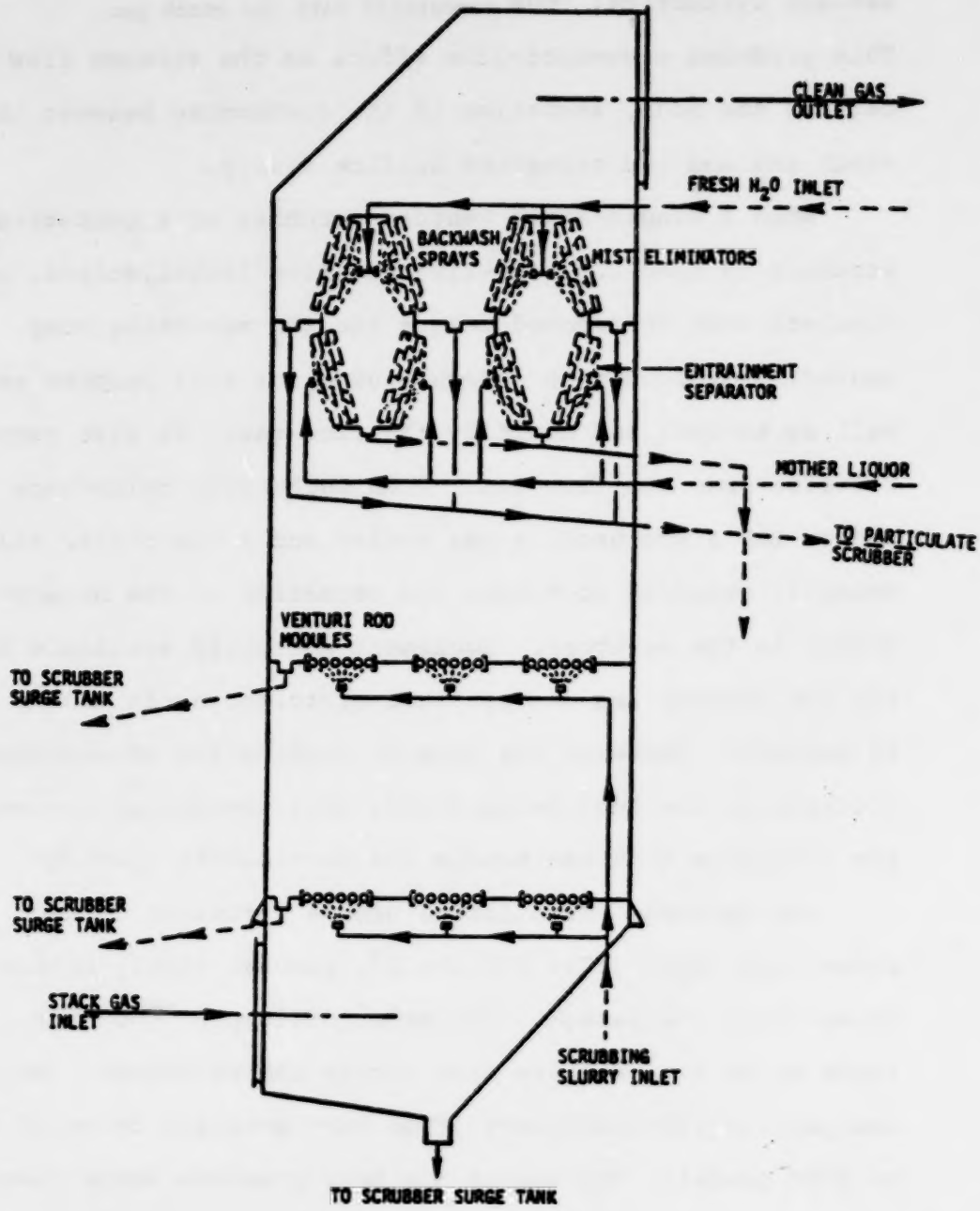

Figure 4. Venturi-rod scrubber.

between cylindrical rods concurrent with the stack gas. This produces a venturi-like effect as the streams flow between the rods, assisting in the contacting between the stack gas and the magnesium sulfite slurry.

When a single-stage venturi scrubber or a venturi-rod scrubber is used for coal-fired utility installations, particulate must be removed before the SO_2 scrubbing step. A venturi prescrubber is commonly used for this purpose as well as to cool and humidify the flue gas. It also removes chloride from the flue gas. When mechanical collectors and/or ESP's are used, a gas cooler and a humidifier are normally required to insure the formation of the hexahydrates in the scrubber. Equipment currently available for fly ash removal has demonstrated efficiencies in excess of 99 percent. Whatever the type or combination of equipment dictated by the fuel being fired, well-engineered systems are available that can handle the particulate loading.

One Japanese installation uses a turbulent contact absorption tower (TCA) for the SO_2 removal step. Little is known about the design. The major difference, however, seems to be the pressure drop across the scrubbers. Well-designed venturi scrubbers often have pressure drops of 4500 to 5000 pascals; TCA towers may have pressure drops ranging from 2500 to 3000 pascals. An SO_2 removal efficiency of 99.8 percent has been claimed for the TCA tower.

As for calcining, it is carried out conventionally in a direct-fired rotary kiln, where coke is added to decompose $MgSO_4$. At the off-site regeneration facility of the Eddystone Generating Station, an oil-fired, fluidized-bed calciner is being investigated. Figure 5 presents a flow diagram of this MgO regeneration system.

4.2 SOLUTIONS TO PROBLEM AREAS

Problems have been primarily mechanical in nature, but some misunderstanding of magnesium oxide characteristics, such as slaking properties, also occurred. When regenerated MgO was first used, it was found to be less reactive than virgin material. High-speed agitation and heating with steam spargers in a slaker proved to be the most effective way to restore the reactivity of regenerated MgO. Heating the MgO above 82°C must be avoided, however, to prevent scaling problems.

Corrosion and erosion have occurred in both the particulate and the SO_2 scrubber systems. The best solution thus far seems to be the use of rubber-lined pipe, pumps, and valves. Spares also should be installed for critical components.

Inside the scrubber, the use of 316L stainless steel, flaked-glass, and corrosion-resistant polyester-resin coatings have given excellent protection; however, some

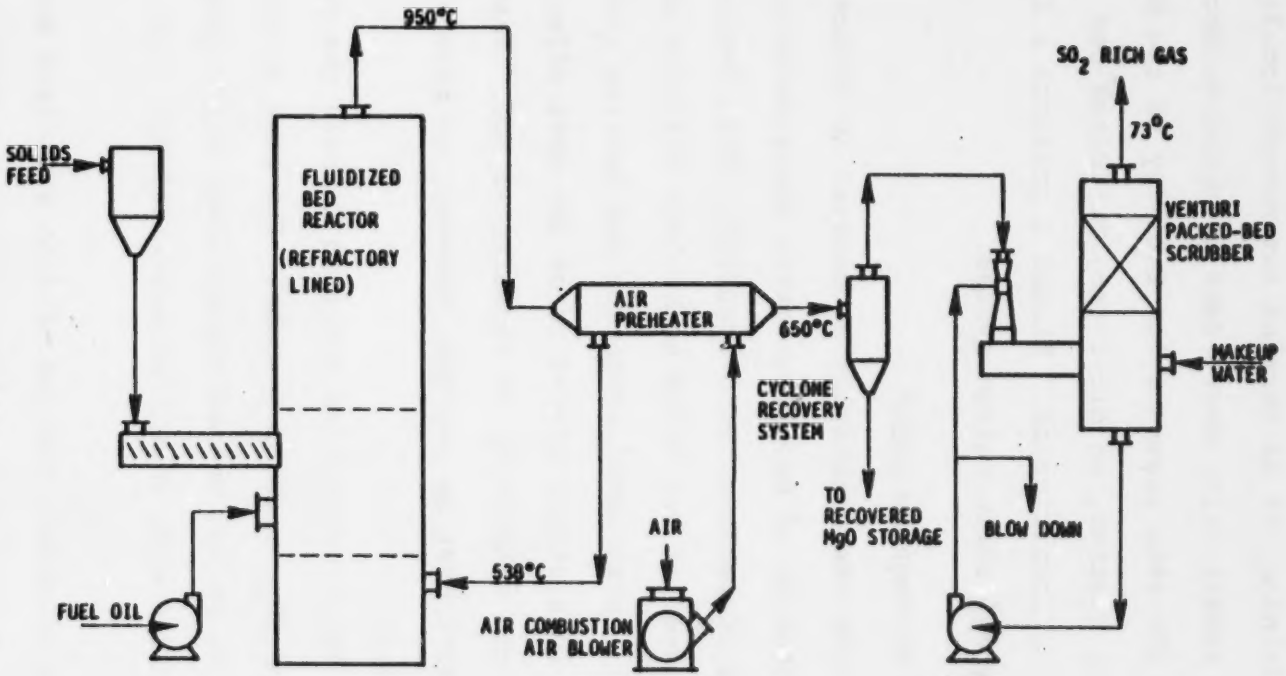

Figure 5. MgO regeneration plant.

erosion/abrasion occurred. In some cases where 316L stainless steel was attacked, Inconel 625 was substituted and showed no sign of corrosion. With these materials of construction, no problems have occurred involving scaling or buildup inside the scrubbers.

The failure or limited capacity of the solids handling equipment often caused plant outages. For example, centrifuge failures have occurred because of excessive wear of the internals.

Some considerations for improving centrifuge design are as follows:

(1) Moving parts should be given stellite hard surfacing treatment to lessen/eliminate wear problems.

(2) Wash-out connections should be installed that are capable of thoroughly washing the centrifuge internals.

(3) The centrifuge should be designed to handle a wide range of solids concentrations and viscosities and provide good separation.

(4) The discharge hopper should be designed to minimize product accumulation.

A capacity problem can be solved by building in excess capacity on conveyors and bucket elevators for transporting dried magnesium sulfite to storage, thus eliminating product surges that can cause conveyor overloading.

Flow in the magnesium sulfite dryer should be concurrent, and internal chain sections and/or external rappers

should be included to dislodge product buildup. The dryer off-gas system should include a cyclone separator for removal of entrained $MgSO_3$, and the cleaned gas stream should then be piped into the SO_2 scrubber for further removal of $MgSO_3$.

Chlorides in the scrubbing solution occur primarily from the combustion of chloride bearing fuel and, to a lesser extent, from chloride in the raw water. The chlorides will react with magnesium oxide and build up in the liquid recirculation loop causing potential problems due to corrosion, decreased SO_2 absorption, and increased magnesium oxide makeup. These problems can be reduced by 1) using a prescrubber and keeping that liquid stream separate from the SO_2 absorber streams, 2) designing for chloride corrosion resistance, and 3) incorporating a purge system into the recirculation loop. Depending on wastewater discharge regulations, the purge stream may require further treatment.

5.0 COST

Costs have been escalated using the Chemical Engineering plant cost index (1957-59 = 100, 1976 = 187.1) to reflect a January 1976 initiation of plant construction. Table 3 presents available cost data for questionnaires and estimates. Because available data were very limited, some annual costs were estimated. Reported data were used where available, as in the case of electricity and MgO, otherwise the following costs were assumed:

Electricity	25 mills/kWh
Magnesium oxide	$204/1000 kg
Labor	$10/h

The life of FGD equipment was assumed to be 20 years; interim replacement, taxes, insurance, and capital charges totaled 13.65 percent of the total 1976 capital cost.

Table 3. COST DATA OF INDIVIDUAL INSTALLATIONS FOR THE MAGNESIUM OXIDE SCRUBBING FGD SYSTEM

Company Name Plant Name Location	Boston Edison Mystic Everett, MS., USA	Potomac Elec. Dickerson Dickerson, MD., USA	Philadelphia Elec. Eddystone Philadelphia, PA., USA	Idemitsu Kosan Co. Chiba Refinery Ichihara, Japan
1. Reported capital costs, $	6,169,000[a,b,c]	6,500,000[a,c]	10,800,000	16,819,000[a,b,c]
Year	1974	1973	1976	1974
Cost inflated to 1976	6,975,000[a,b]	8,436,000[a]	10,800,000[a,b,c]	18,723,000[a,b]
$/kW	45[a,b]	88.8[a]	90[a,b,c]	102.5[a,b]
2. Reported annual costs, $				
Total fixed costs	1,301,000	1,701,000[d]	2,014,000[d]	3,492,000
Fixed charge factor, %[f]	18.65	20.16	18.65	18.65
Total variable costs	1,250,000[d,c]	1,712,000[c,e,d,g]	2,323,000[d,g]	
3. Boiler operating hours	4300	7993/1974	6549/1974	7920
Capacity factor	0.50	0.81	0.68	
Generated, MWh	678,900	674,082	714,816	
4. Reported cost per kWh, mills/kWh	3.76[d]	5.11[c]	6.07[d]	8.94[c]
Fixed cost (boiler operation)	1.92	2.52	2.82	2.76[h]
Variable costs	1.84	2.54	3.25	
Total cost, mills/kWh[i]				
4000 h	3.97	6.73	7.15	
8000 h	2.92	4.49	5.05	

[a] Scrubbing unit.
[b] Regeneration unit (acid plant not included).
[c] Based on questionnaire.
[d] Estimation.
[e] Ref. 13. Bibliography.
[f] Fixed costs normally based on 7-year depreciation in Japan.
[g] No product credit.
[h] Equivalent cost, boiler is not an electric utility boiler.
[i] Based on load of 100% for hours indicated.

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APPENDIX A

APPENDIX A

QUESTIONNAIRE DATA ON THE MAGNESIUM OXIDE FLUE GAS DESULFURIZATION PROCESS

Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
1.	GENERAL DESCRIPTION		
1.1	Company	Boston Edison	Philadelphia Electric Co.
1.2	Plant name	Mystic No. 6	Eddystone Station
1.3	Preparer's name and mailing address	Michael D. Giordano 11499 Chester Rd. Cincinnati, Ohio 45246	Michael D. Giordano 11499 Chester Rd. Cincinnati, Ohio 45246
1.4	Principal plant products	Electric power	Electric power
	Brief description of area surrounding plant site and factors for choice of FGD system	N.A.	N.A.
1.5	Plant location	Everett, Massachusetts	Eddystone, Pennsylvania
1.6	STATION CAPACITY		
1.6.1	Total boiler generating capacity, MW /number of boilers	1065/4	1370/4
1.6.2	Capacity served by FGD system, MW	155	308
	amount of flue gas treated, Nm^3/h	N.A.	N.A.
	% of boiler flue gas treated by FGD system	100	33
	Number of boilers served	1	1

N.A. - Not available.

Nm^3/h - m^3/h (0°C and 101.325kPa)

(continued)

APPENDIX A

QUESTIONNAIRE DATA ON THE MAGNESIUM OXIDE FLUE GAS DESULFURIZATION PROCESS

Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
1.	GENERAL DESCRIPTION		
1.1	Company	Potomac Electric Power Co.	Idemitsu Kosan Co., Ltd.
1.2	Plant name	Dickerson	Chiba Refinery
1.3	Preparer's name and mailing address	Michael D. Giordano 11499 Chester Rd. Cincinnati, Ohio 45246	Jumpei Ando Chuo University 1-13-27, Kasuga, Bunkyo-Ku Japan
1.4	Principal plant products	Electric power	Petroleum Products
	Brief description of area surrounding plant site and factors for choice of FGD system	N.A.	Faces Tokyo Bay near Ichihara and Chiba
1.5	Plant location	Dickerson, Maryland	Ichihara, Chiba Prefecture
1.6	STATION CAPACITY		
1.6.1	Total boiler generating capacity, MW/number of boilers	570/3	1627/9
1.6.2	Capacity served by FGD system, MW	95	293
	amount of flue gas treated, Nm^3/h	N.A.	N.A.
	% of boiler flue gas treated by FGD system	50 1	47% of total 100% of two
	Number of boilers served	1	2

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
1.7	TYPE OF FGD SYSTEM	Magnesium oxide regenerative scrubbing	MgO Scrubbing
1.8	Number of scrubber trains in parallel	1	3
1.9	FGD system installed on new or existing boiler	Existing	Existing
1.10	STATUS OF FGD SYSTEM operational or under construction, date started	Shutdown, 1972	Operational, 1975
2.	BOILER DATA		
2.1	Boiler identification number	Mystic No. 6	Eddystone No. 1
2.2	Maximum continuous heat input, MW	415	853
2.3	Max. cont. net generating capacity, MW	148	308
2.4	Flue gas flow rate, Nm^3/h temperature at induced draft fan, °C	643,000 149	1,027,700 149
2.5	Boiler manufacturer	Combustion Engineering	Combustion Engineering
2.5.1	Fuel type	Oil	Coal
2.5.2	Firing configuration	Tangential firing	Tangential
2.6	Year placed in service	1961	1959
2.7	Type service	Cycling units	Base load
2.8	Stack height, m	99	76
2.8.1	Flue gas exit velocity, m/s	21	N.A.
2.9	Average boiler operation, h/yr	4300	6549/1974
2.10	Average boiler capacity factor	0.50	0.68/1974

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
1.7	TYPE OF FGD SYSTEM	MgO scrubbing	Magnesium scrubbing
1.8	Number of scrubber trains in parallel	1	1
1.9	FGD system installed on new or existing boiler	Existing	Existing
1.10	STATUS OF FGD SYSTEM operational or under construction, date started	Shutdown, September 1973	Operational August 1974
2.	BOILER DATA		
2.1	Boiler identification number	3	No. 5 and No. 6
2.2	Maximum continuous heat input, MW	425	N.A.
2.3	Max. cont. net generating capacity, MW	190	186
2.4	Flue gas flow rate, Nm^3/h temperature at induced draft fan, °C	686,000 126	460,000 N.A.
2.5	Boiler manufacturer	Combustion Engineering	N.A.
2.5.1	Fuel type	Coal	Heavy oil C
2.5.2	Firing configuration	Pulverized	Corner firing
2.6	Year placed in service	1962	1972
2.7	Type service	Base load	Base load
2.8	Stack height, m	122	180
2.8.1	Flue gas exit velocity, m/s		N.A.
2.9	Average boiler operation, h/yr	7993/1974	7920/1976
2.10	Average boiler capacity factor	0.81	N.A.

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
2.11	COAL COMPOSITION	N.A.	
2.11.1	Lower heating value, kJ/kg	N.A.	28,000
2.11.2	Sulfur content, %		2.3
2.11.3	Ash content, %		12.0
2.11.4	Chloride content, %		N.A
2.11.5	Moisture, %		5-8
2.12	OIL COMPOSITION		N.A.
2.12.1	Grade/heating value, GJ/m^3	No. 6/41.5	
2.12.2	Sulfur content, %	2.0	
2.12.3	Ash content, %	0.07	
2.12.4	Chloride content, %	N.A.	
2.13	PARTICULATE REMOVAL EQUIPMENT NOT PART OF FGD SYSTEM	NONE	
2.13.1	Type		Mechanical and electrostatic precipitators
2.13.2	Manufacturer		American Standard and Western Precip.
2.13.3	Efficiency, % removal		95
2.13.4	Emission rate, g/s		75.6

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
2.11	COAL COMPOSITION		N.A.
2.11.1	Lower heating value, kJ/kg	27000	
2.11.2	Sulfur content, %	2.2	
2.11.3	Ash content, %	14.1	
2.11.4	Chloride content, %	N.A.	
2.11.5	Moisture, %	N.A.	
2.12	OIL COMPOSITION	N.A.	
2.12.1	Grade/heating value, kJ/kg		
2.12.2	Sulfur content, %		1.8
2.12.3	Ash content, %		0.04
2.12.4	Chloride content, % ppm		10
2.13	PARTICULATE REMOVAL EQUIPMENT NOT PART OF FGD SYSTEM		
2.13.1	Type	Electrostatic precipitator	Electrostatic precipitator
2.13.2	Manufacturer	Research-Cottrell	Nihonkai Juko KK
2.13.3	Efficiency, % removal	94	80
2.13.4	Emission [illegible]	N.A.	N.A.

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
3.	EMISSION REGULATIONS		
3.1	ATMOSPHERIC EMISSIONS		
3.1.1	SO_2	State currently requires 1% sulfur oil as fuel	0.258 ng/kJ
3.1.2	Particulate, ng/kJ	0.052	0.043
3.1.3	NO_x	None	N.A.
3.1.4	Other	N.A.	N.A.
3.1.5	Legal basis	State regulations	Local regulations
3.2	RESTRICTIONS APPLICABLE TO LIQUID OR SOLID WASTES, LEGAL BASIS	N.A.	
4.	FGD SYSTEM INFORMATION		
4.1	Process name	Magnesium oxide regenerative scrubbing	Magnesium base wet scrubbing
4.1.1	Supplier or design company	CHEMICO	United Engineers and Constructors
4.1.2	WHY WAS THIS SYSTEM CHOSEN?	N.A.	No area available for sludge disposal
4.1.3	PROCESS DESCRIPTION	Appendix Figures A, B, and C	Appendix Figure D

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
3.	EMISSION REGULATIONS	N.A.	
3.1	ATMOSPHERIC EMISSIONS		
3.1.1	SO_2, Nm^3/h		183
3.1.2	Particulate, mg/Nm^3		30
3.1.3	NO_x, ppm		130
3.1.4	Other		N.A.
3.1.5	Legal basis		Local government
3.2	RESTRICTIONS APPLICABLE TO LIQUID OR SOLID WASTES, LEGAL BASIS		N.A.
4.	FGD SYSTEM INFORMATION		
4.1	Process name	Magnesium oxide	Magnesium scrubbing
4.1.1	Supplier or design company	CHEMICO	Mitsui Miike Machinery
4.1.2	WHY WAS THIS SYSTEM CHOSEN?	N.A.	High SO_2 removal efficiency. Less wastewater. Useful by-product (elemental sulfur)
4.1.3	PROCESS DESCRIPTION	N.A.	Appendix Figure E

(continued)

APPENDIX A (continued) Part A

	CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.2 SO_2 SCRUBBING AGENT		
4.2.1 Name	Magnesium oxide	Magnesium oxide
4.2.2 Chemical composition	MgO CaO SiO_2 SO_4	MgO
Components, % by weight	MgO, 97.6 CaO, 1.5 SiO_2, 0.5 SO_4, 10.9	MgO, 97.5
4.2.3 Stoichiometric requirement	N.A.	1.0
4.2.4 Addition rate, kg/h	N.A.	580
4.2.5 Particle size, mm	N.A.	N.A.
4.3 REGENERATION AGENT	None	None
4.3.1 Name		
4.3.2 Composition		
4.3.3 Stoichiometric requirement		
4.3.4 Addition rate		
4.3.5 Particle size		
4.4 SALABLE PRODUCT		
4.4.1 Name	N.A.	N.A.
4.4.2 Chemical composition	N.A.	SO_2 to be sold to sulfuric acid plant SO_2, 10-20 H_2, CO_2, H_2O, etc., 80-90
Components, % by weight		
4.4.3 Amount generated/year, Mg/yr	4540 (over 2 year test program)	8070
4.4.4 Sales potential	High quality (93%) commercial grade H_2SO_4	Good

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.2	SO_2 SCRUBBING AGENT		
4.2.1	Name	Magnesium oxide slurry	Magnesium hydroxide
4.2.2	Chemical composition	MgO	$Mg(OH)_2$ CaO
	Components, % by weight	MgO,90	$Mg(OH)_2$,97.3 CaO/0.8
4.2.3	Stoichiometric requirement	1.5	1
4.2.4	Addition rate, tons/yr	N.A.	362
4.2.5	Particle size, mm	N.A.	Less than 0.044
4.3	REGENERATION AGENT	None	None
4.3.1	Name		
4.3.2	Composition		
4.3.3	Stoichiometric requirement		
4.3.4	Addition rate		
4.3.5	Particle size		
4.4	SALABLE PRODUCT	H_2SO_4	Elemental sulfur
4.4.1	Name	N.A.	N.A.
4.4.2	Chemical composition Components, % by weight	H_2SO_4,93	S,99.9
4.4.3	Amount generated/year	0.76 kg/s (dry)	15,000 Mg/yr
4.4.4	Sales potential	Good	N.A.

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.4.5	BY-PRODUCT RECOVERY PROCESS.	N.A.	
	Name		$MgSO_3$ roasting
	Vendor		Copeland System, Inc.
	Process description		Calcining of magnesium
4.4.6	Intermediate production composition		Sulfite to regen. MgO
4.5	BY-PRODUCT NO. 2	None	None
4.6	WASTE STREAMS	None	None
4.6.1	Type of waste		
	% water		
4.6.2	Amount generated in one hour at full load		
4.6.3	Disposal method		
4.6.4	Disposal site and distance from plant, km		
4.6.5	Potential effects on environment from sludge disposal		
4.7	SLUDGE FIXATION PROCESS	None	None

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.4.5	BY-PRODUCT RECOVERY PROCESS		
	Name	N.A.	N.A.
	Vendor	Essex Chemical Co.	Claus Furnace, Japan Corp.
	Process description	Calcine $MgSO_3$ to recover SO_2 and MgO	N.A.
4.4.a	Intermediate production composition	N.A.	N.A.
4.5	BY-PRODUCT NUMBER 2	None	None
4.6	WASTE STREAMS	None	Wastewater
4.6.1	Type of waste		
	E water		N.A.
4.6.2	Amount generated in one hour at full load, Mg/yr		0.09
4.6.3	Disposal method		N.A.
4.6.4	Disposal site and distance from plant, km		N.A.
4.6.5	Potential effects on environment from sludge disposal		N.A.
4.7	SLUDGE FIXATION PROCESS	None	None

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.8	FGD EQUIPMENT DATA		
4.8.1	FIRST SCRUBBER		
4.8.1.1	Type	Venturi	Venturi (rod type)
4.8.1.2	L/G[6] ratio, ℓ/Nm^3	Inlet - 13.9 Outlet - 16.6	2.4
4.8.1.3	Gas velocity through absorber or scrubber, m/s	N.A.	N.A.
4.8.1.4	Absorbent (adsorbent) flow, m^3/h	3420	288
4.8.1.5	Gas flow entering scrubber, Nm^3/h	244,800	367,200
4.8.1.6	Gas temperature, °C	138	145
4.8.1.7	Gas composition		N.A.
	SO_2 in → out, ppm	1042 → 89	
	% removal	≥90	
	Particulate in → out, kg/s	0.50 → 0.20	
	% removal	≥60	
	NO_x in → out, ppm	N.A.	
	CO_2 in → out, ppm	N.A.	
	Other in → out	N.A.	
4.8.1.8	Material of construction	Carbon steel	316L S.S.
4.8.1.9	Type of lining	Polyester (Neil Rigiflake)	N.A.
4.8.1.10	Adsorbent and sand losses	N.A.	N.A.
4.8.1.11	Number of stages	1	1
4.8.1.12	Packing type and composition	N.A.	N.A.

[6]Liquid-to-gas ratio

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.8	FCD EQUIPMENT DATA		
4.8.1	FIRST SCRUBBER		None
4.8.1.1	Type	Venturi	
4.8.1.2	L/G ratio, l/Nm^3	2.67	
4.8.1.3	Gas velocity through absorber or scrubber, m/s	N.A.	
4.8.1.4	Absorbent (adsorbent) flow, m^3/h	N.A.	
4.8.1.5	Gas flow entering scrubber, Nm^3/h	6120	
4.8.1.6	Gas temperature, °C	126	
4.8.1.7	Gas composition, ppm		
	SO_2 in + out	N.A.	
	% removal	N.A.	
	Particulate in + out, mg/m^3	N.A.	
	% removal	N.A.	
	NO_x in + out, ppm	Not measured	
	CO_2 in + out	N.A.	
	Other in + out	N.A.	
4.8.1.8	Material of construction	Steel shell	
4.8.1.9	Type of lining	Fiberglass reinforced plastic (FRP)	
4.8.1.10	Adsorbent and sand losses	N.A.	
4.8.1.11	Number of stages	1	
4.8.1.12	Packing type and composition	None	

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.8.2	SECOND SCRUBBER	None	
4.8.2.1	Type		Venturi (rod type)
4.8.2.2	L/G ratio, l/Nm^3		10-12
4.8.2.3	Gas velocity through absorber or scrubber, m/s		N.A.
4.8.2.4	Absorbent (adsorbent) flow, m^3/h		5.04
4.8.2.5	Gas flow entering scrubber, Nm^3/h		6840
4.8.2.6	Gas temperature, °C		53
4.8.2.7	Gas composition		N.A.
	SO_2 in + out, ppm		
	% removal		
	Particulate in + out, mg/Nm^3		
	% removal		
	NO_x in + out, ppm		
	% removal		
	CO_2		
	Other		
4.8.2.8	Material of construction		Lined carbon steel
4.8.2.9	Type of lining		Polyurethane
4.8.2.10	Number of stages		2
4.8.2.11	Packing type and composition		N.A.

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.8.2	SECOND SCRUBBER		
4.8.2.1	Type	Venturi	Venturi (Chemico type)
4.8.2.2	L/G ratio, ℓ/Nm^3	5.33	N.A.
4.8.2.3	Gas velocity through absorber or scrubber, m/s	N.A.	N.A.
4.8.2.4	Absorbent (adsorbent) flow, m^3/h	N.A.	N.A.
4.8.2.5	Gas flow entering scrubber, Nm^3/h	6120	460,000
4.8.2.6	Gas temperature, °C	52	185
4.8.2.7	Gas composition		
	SO_2 in → out	1.3 kg/GJ → 0.13 kg/GJ	2850 ppm → Below 120[a]
	% removal	90	Over 95
	Particulate in → out mg/Nm^3	N.A.	130 → 30-60
	% removal	N.A.	54-77
	NO_x in → out, ppm	N.A.	110 → 110
	% removal	N.A.	11.4
	CO_2	N.A.	N.A.
	Other	N.A.	N.A.
4.8.2.8	Material of construction	Steel shell	N.A.
4.8.2.9	Type of lining	FRP	N.A.
4.8.2.10	Number of stages	1	2
4.8.2.11	Packing type and composition	N.A.	N.A.

[a] Overall effect of 1st and 2nd stages of venturi scrubber.

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.8.3	CLEAR WATER TRAY	N.A.	N.A.
4.8.3.1	Type		
4.8.3.2	L/G ratio, l/Nm^3		
4.8.3.3	Source of water		
4.8.4	DEMISTER		
4.8.4.1	Type	Chevron	Chevron
4.8.4.2	Number of passes (stages)	6	2
4.8.4.3	Material of construction	Polypropylene	Noryl Plastic
4.8.4.4	Method of cleaning	Intermittently cleaned	Clean liquor and water spray
4.8.4.5	Flow rate during cleanings, Nm^3/h	N.A.	140 (first stage) 90 (second stage)
4.8.4.6	Frequency and duration of routine cleaning	N.A.	Continuous (first stage) As needed (second stage)
4.8.5	REHEATER	None	
4.8.5.1	Type		Direct
4.8.5.2	Heat load, kW		12,000
4.8.5.3	Heat transfer surface area, m^2		N.A.
4.8.5.4	Temperature of gas in, °C out, °C		53 110
4.8.5.5	Reheater materials of construction		N.A.

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.8.3	CLEAR WATER TRAY	None	N.A.
4.8.3.1	Type		
4.8.3.2	L/G ratio, ℓ/Nm^3		
4.8.3.3	Source of water		
4.8.4	DEMISTER		
4.8.4.1	Type	Baffle	Chemico type
4.8.4.2	Number of passes (stages)	3	N.A.
4.8.4.3	Material of construction	FRP	N.A.
4.8.4.4	Method of cleaning	Up-spray	N.A.
4.8.4.5	Flow rate during cleanings, Nm^3/h	N.A.	N.A.
4.8.4.6	Frequency and duration of routine cleaning	Once every 8 h for 5-10 min	N.A.
4.8.5	REHEATER	N.A.	
4.8.5.1	Type		N.A.
4.8.5.2	Heat load, kW		N.A.
4.8.5.3	Heat transfer surface area, m^2		N.A.
4.8.5.4	Temperature of gas in, °C out, °C		N.A. N.A.
4.8.5.5	Reheater materials of construction		Refractory-lined carbon steel

(continued)

APPENDIX A (continued) Part A

		CHEMICO Hyatic No. 6	United Engineers and Constructors Eddystone Station
4.8.6	SCRUBBER TRAIN PRESSURE DROP, mbar		
4.8.6.1	First scrubber	17.5-20	10-42
4.8.6.2	Second scrubber	None	30
4.8.6.3	Clear water tray	N.A.	N.A.
4.8.6.4	Demister	5-7.5	2.5
4.8.6.5	Reheater	None	None
4.8.6.6	Ductwork	N.A.	10
4.8.6.7	Total FGD system	22.5-27.5	73-85
4.8.6.8	I.D. fan	30	
4.9	SYSTEM MAKEUP WATER ADDED, m^3/h	N.A.	0.96
4.9.1	Water added per mole of SO_2 removed, m^3/kg mol		N.A.
4.9.2	Source of makeup water		River
4.9.3	Where is water added		Particulate scrubber surge tank
4.10	BYPASS SYSTEM		
4.10.1	Can flue gas be bypassed?	Yes	Yes
4.10.2	Height of gas discharge when bypassed, m	N.A.	76
4.11	SLURRY DATA, pH, % solids by weight		
4.11.1	Fresh slurry to scrubber	≈10,8.6	10.8-10
4.11.2	Recirculating slurry	7[a],14	1-3.1
4.11.3	Waste slurry	N.A.	6-7.8-10

[a] Slurry recycle pH affects SO2 outlet concentrations linearly: ranging from pH 6.0 ([SO_2] = 75 ppm).

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.8.6	SCRUBBER TRAIN PRESSURE DROP, mbar		
4.8.6.1	First scrubber	27.4	N.A.
4.8.6.2	Second scrubber	14.9-29.9	N.A.
4.8.6.3	Clear water tray	2.5	N.A.
4.8.6.4	Demister	N.A.	N.A.
4.8.6.5	Reheater	N.A.	N.A.
4.8.6.6	Ductwork	25	N.A.
4.8.6.7	Total FGD system	87	50
4.8.6.8	I.D. fan	N.A.	N.A.
4.9	SYSTEM MAKE UP WATER ADDED, m^3/h	0.378 (first stage) 0.048 (second stage)	N.A.
4.9.1	Water added per mole of SO_2 removed, m^3/kg mol	N.A.	N.A.
4.9.2	Source of make-up water	N.A.	River
4.9.3	Where is water added	First and second stages	Scrubber
4.10	BYPASS SYSTEM		
4.10.1	Can flue gas be bypassed?	Yes to stack	Yes
4.11	SLURRY DATA, pH, % solids by weight	N.A.	N.A.
4.11.1	Fresh slurry to scrubber		
4.11.2	Recirculating slurry		
4.11.3	Waste slurry		

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.12	SPENT SLURRY SYSTEM		
4.12.1	Clarifiers	N.A.	
4.12.1.1	Number		1
4.12.1.2	Diameter, m		12.2
4.12.2	Solid separation		N.A.
4.12.2.1	Type	Centrifuge (solid bowl)	N.A.
4.12.2.2	Cloth area, m^2	N.A.	N.A.
4.12.2.3	Capacity (wet cake), kg/S	1.6	N.A.
4.13	SPACE REQUIRED, EXCLUSIVE OF SLUDGE DISPOSAL AREA, m^2	~1900 (scrubber only)	N.A.
4.14	PERSONNEL, man-hours/yr		N.A.
4.14.1	Operating	a	
4.14.1.1	Specialists	N.A.	
4.14.1.2	Unskilled	N.A.	
4.14.2	Maintenance and repair	N.A.	
4.15	ENVIRONMENTAL LOAD		N.A.
4.15.1	Wastewater treatment process	N.A.	
4.15.2	Concentrations	N.A.	
4.15.3	Appearance of flue gas discharge - any mist problems?	Dense vapor plume No mist problems	

[a] Total O&M cost over 2-year test program for both scrubber and calciner = $4,205,799.

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.12	SPENT SLURRY SYSTEM	N.A.	N.A.
4.12.1	Clarifiers		
4.12.1.1	Number		
4.12.1.2	Diameter, m		
4.12.2	Solid separation		
4.12.2.1	Type		
4.12.2.2	Cloth area, m^2		
4.12.2.3	Capacity (wet cake), tons/h		
4.13	SPACE REQUIRED, EXCLUSIVE OF SLUDGE DISPOSAL AREA, m^2	N.A.	N.A.
4.14	PERSONNEL		
4.14.1	Operating	N.A.	N.A.
4.14.1.1	Specialists	8736 (operator)	4 persons/shift
4.14.1.2	Unskilled	8736 (helper)	1 person/shift
4.14.2	Maintenance and repair	3000	33,120 h/yr
4.15	ENVIRONMENTAL LOAD	N.A.	
4.15.1	Wastewater treatment process		N.A.
4.15.2	Concentrations		N.A.
4.15.3	Appearance of flue gas discharge - any mist problems?		Steam Yes, mist problems once or twice a year

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
4.16	FGD SYSTEM OPERATION		N A.
4.16.1	% of boiler load FGD can operate	100	
4.16.2	Lag time	No lag time observed	
5.	ACTUAL COSTS FOR TOTAL FGD INSTALLATION, year	1974	1975
5.1	CAPITAL COSTS, $		
5.1.1	Scrubber	5,250,742	50%[a]
5.1.2	Absorbent/regeneration system	N.A.	7%
5.1.3	Sludge disposal system	N.A.	N.A.
5.1.4	Site improvements	916,588	2%
5.1.5	Land, roads, tracks, substation	N.A.	12%
5.1.6	Engineering costs	Included in above	1%
5.1.7	Contractor's fee	N.A.	8%
5.1.8	Interest during construction	N.A.	10%
	Construction period, mo	14	N.A.
5.1.9	License fees	N.A.	N.A.
5.1.10	Other	N.A.	N.A.
5.1.11	Total, $	3,635,000 (Mystic) 2,532,330 (Rumford MgO regen. facility)	20,273,000 (Includes three particulate scrubbers)

[a] Percentages of total installed capital cost. Total scrubber cost $3,034,000.

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
4.16	FGD SYSTEM OPERATION		
4.16.1	% of boiler load FGD can operate	4:1 turndown	26-100
4.16.2	Lag time	N.A.	N.A.
5.	ACTUAL COSTS FOR TOTAL FGD INSTALLATION year	1974	1974
5.1	CAPITAL COSTS, $	N.A.	N.A.
5.1.1	Scrubber	N.A.	N.A.
5.1.2	Absorbent/regeneration system	N.A.	N.A.
5.1.3	Sludge disposal system	N.A.	N.A.
5.1.4	Site improvements	N.A.	N.A.
5.1.5	Land, roads, tracks, substation	N.A.	N.A.
5.1.6	Engineering costs	N.A.	N.A.
5.1.7	Contractor's fee	N.A.	N.A.
5.1.8	Interest during construction	N.A.	N.A.
	Construction period, mo	24	N.A.
5.1.9	License fees	N.A.	N.A.
5.1.10	Other	N.A.	N.A.
5.1.11	Total, $ $/kW	6,500,000 (for 95 MW) 68 (excluding regeneration equipment)	14,500,000 (4.35 billion yen) N.A. (excluding Claus furnace)

(continued)

APPENDIX A (continued) Part A

		CHEMICO Nystic No. 6	United Engineers and Constructors Eddystone Station
5.2	ANNUAL COSTS, $		
5.2.1	Interest	N.A.	N.A.
5.2.2	Depreciation	N.A.	N.A.
5.2.3	Insurance and taxes	N.A.	N.A.
5.2.4	Total fixed costs	N.A.	N.A.
5.2.5	Labor and overhead Average man-hour cost cost/year	N.A.	N.A.
5.2.6	Raw material breakdown Unit cost Annual cost	N.A.	N.A.
5.2.7	Utilities, unit cost, annual costs	N.A.	
	Electricity		N.A.
	Water		N.A.
	Steam		N.A.
	Fuel		N.A.
	Total		2,400,000
5.2.8	Maintenance cost/year	4,205,799 (oper. and maint. cost over 2 yr)	N.A.
5.2.9	Other costs, list	N.A.	N.A.
5.2.10	Total variable cost	N.A.	N.A.
5.2.11	Total annual cost	N.A.	N.A.
5.3	WASTE DISPOSAL COST, $	N.A.	N.A.
	Unit cost Annual cost		
5.4	BY-PRODUCT CREDIT, $	N.A.	
5.4.1	By-product 1		948,000 (1972)
5.4.2	By-product 2		N.A.

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
5.2	ANNUAL COSTS, $		
5.2.1	Interest	N.A.	N.A.
5.2.2	Depreciation	N.A.	N.A.
5.2.3	Insurance and taxes	N.A.	N.A.
5.2.4	Total fixed costs	15.5% of capital	6,000,000 (1.8 billion yen in 1976 including additional equipment)
5.2.5	Labor and overhead	N.A.	
	Average man-hour cost cost/year		N.A.
5.2.6	Raw material breakdown		N.A.
	Unit cost Annual cost	150/ton of makeup MgO N.A.	
5.2.7	Utilities, unit cost, annual costs	N.A.	N.A.
	Electricity Water Steam Fuel Total		
5.2.8	Maintenance cost/year	N.A.	N.A.
5.2.9	Other costs, list	N.A.	N.A.
5.2.10	Total variable cost/year	500,000 (1974)	2,147,000 (644 million yen in 1976)
5.2.11	Total annual cost	N.A.	8,147,000 (2.444 billion yen in 1976)
5.3	WASTE DISPOSAL COST, $	N.A.	N.A.
	Unit cost Annual cost		
5.4	BY-PRODUCT CREDIT, $	None	None
5.4.1	By-product 1		
5.4.2	By-product 2		

(continued)

APPENDIX A (continued) Part A

		CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
5.5	NET ANNUAL COST, kWh/mills	N.A.	N.A.
5.5.1	Cost last year		
5.5.2	Cost at maximum continuous rating		
5.5.3	Cost without FGD		
6.	ENERGY REQUIREMENTS, kW		N.A.
	Pumps	N.A.	
	Fan	N.A.	
	Reheat	N.A.	
	Reductant chemicals	N.A.	
	Other	N.A.	
	Total	3180 (connected load)	

(continued)

APPENDIX A (continued) Part B

		CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
5.5	NET ANNUAL COST, kWh/mills		N.A.
5.5.1	Cost last year	5 (estimated FGD system cost)	
5.5.2	Cost at maximum continuous rating	N.A.	
5.5.3	Cost without FGD	N.A.	
6.	ENERGY REQUIREMENTS, kW	N.A.	
	Pumps		1960
	Fan		3400
	Reheat		2620
	Reductant chemicals		Coke-170 kg/h
	Other		Steam-1.5 ton/h
	Total		N.A.

(continued)

APPENDIX A (continued) Part A

	CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
7. PERFORMANCE HISTORY: OPERATING PROBLEMS AND SOLUTIONS	[a]	
7.1 Scrubber system	Overheating from steam freezing protection system causes scaling in $Mg(OH)_2$ piping	Extensive corrosion caused by low pH, thermal shock, higher blowdowns, cold start and high chloride concentration
7.2 Demister	N.A.	N.A.
7.3 Reheater	N.A.	N.A.

[a]For complete performance history - see EPA 600/2-75-075, October 1975 for project report.

(continued)

APPENDIX A (continued) Part B

	CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
7. PERFORMANCE HISTORY: OPERATING PROBLEMS AND SOLUTIONS		N.A.
7.1 Scrubber system	N.A.	
7.2 Demister	Accidentally broken during manual cleaning	
7.3 Reheater	N.A.	

(continued)

APPENDIX A (continued) Part A

	CHEMICO Mystic No. 6	United Engineers and Constructors Eddystone Station
7.4 Fan	N.A.	Material failure between shaft and hub
7.5 Sludge system	N.A.	N.A.
7.6 Regeneration system	$MgSO_3$ crystal dryer was poorly designed and operated. Redesigned and fuel changed.	Relocation problem due to H_2SO_4 plant shutdown (6-month delay)
7.7 By-product system	N.A.	N.A.
7.8 Other	Excessive wear on centrifuge parts, erosion of pumps, piping and valves	N.A.
7.9 Availability,%	50 (June 1971 - June 1974)	N.A.

(continued)

APPENDIX A (continued) Part B

	CHEMICO Dickerson	Mitsui Miike Machinery Chiba Refinery
7.4 Fan	Rubber lining of housing with Inconel wheel appears satisfactory	N.A. N.A.
7.5 Sludge system	N.A.	
7.6 Regeneration system	N.A.	Falloff of refractories in a kiln and other problems have been solved
7.7 By-product system	N.A.	N.A.
7.8 Other	By pass dampers bind and leak	N.A.
7.9 Availability, %	N.A.	100 (over the recent 7 month period)

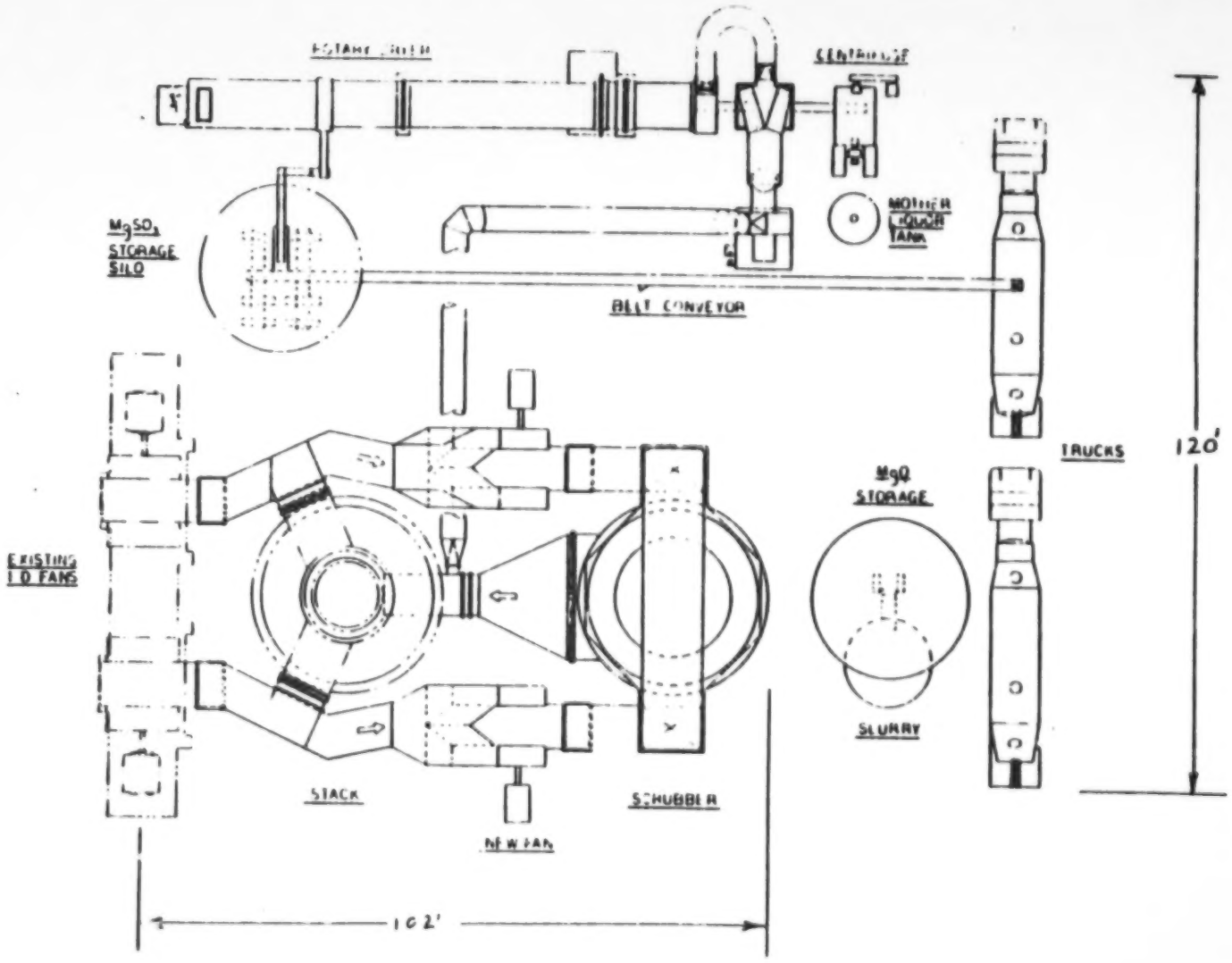

Figure A. General arrangement plan scrubbing system for 155 MW capacity at Mystic No. 6 Station of Boston Edison Company.

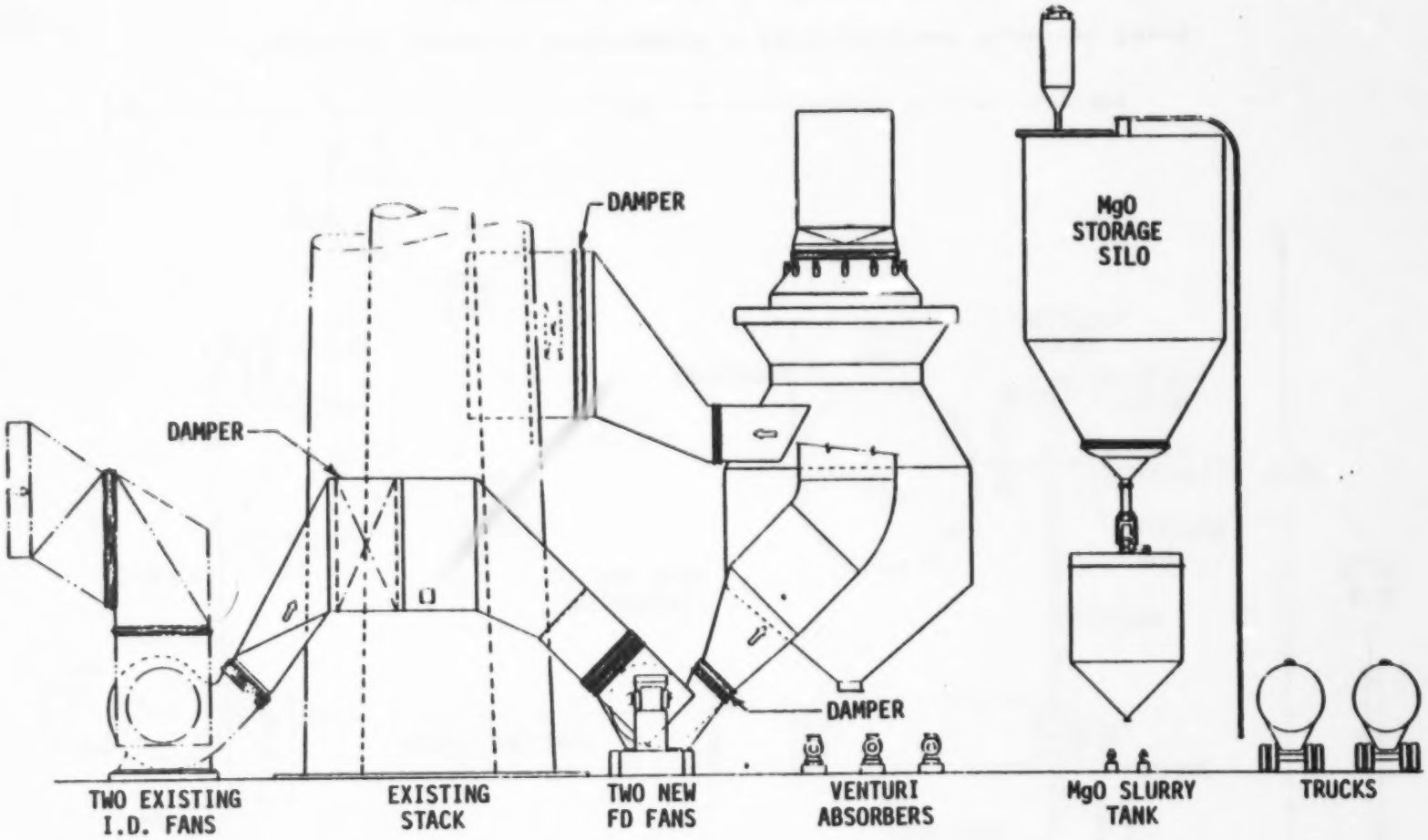

Figure B. Damper location and ducts flue gas desulfurization system Mystic No. 6 Station of Boston Edison Co.

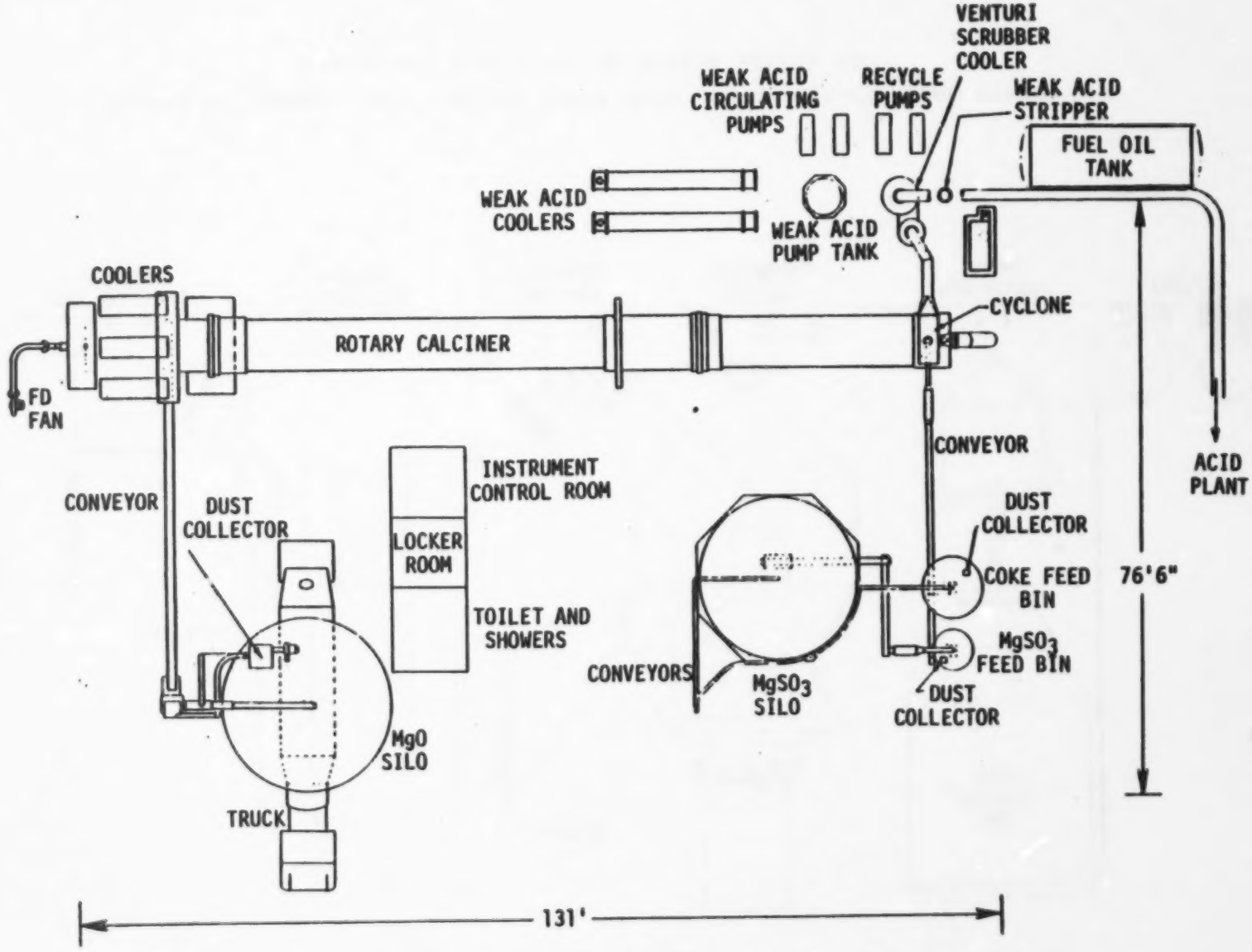

Figure C. General Arrangement - Plan Calciner Plant at Essex Chemical Co. Rumford, Rhode Island

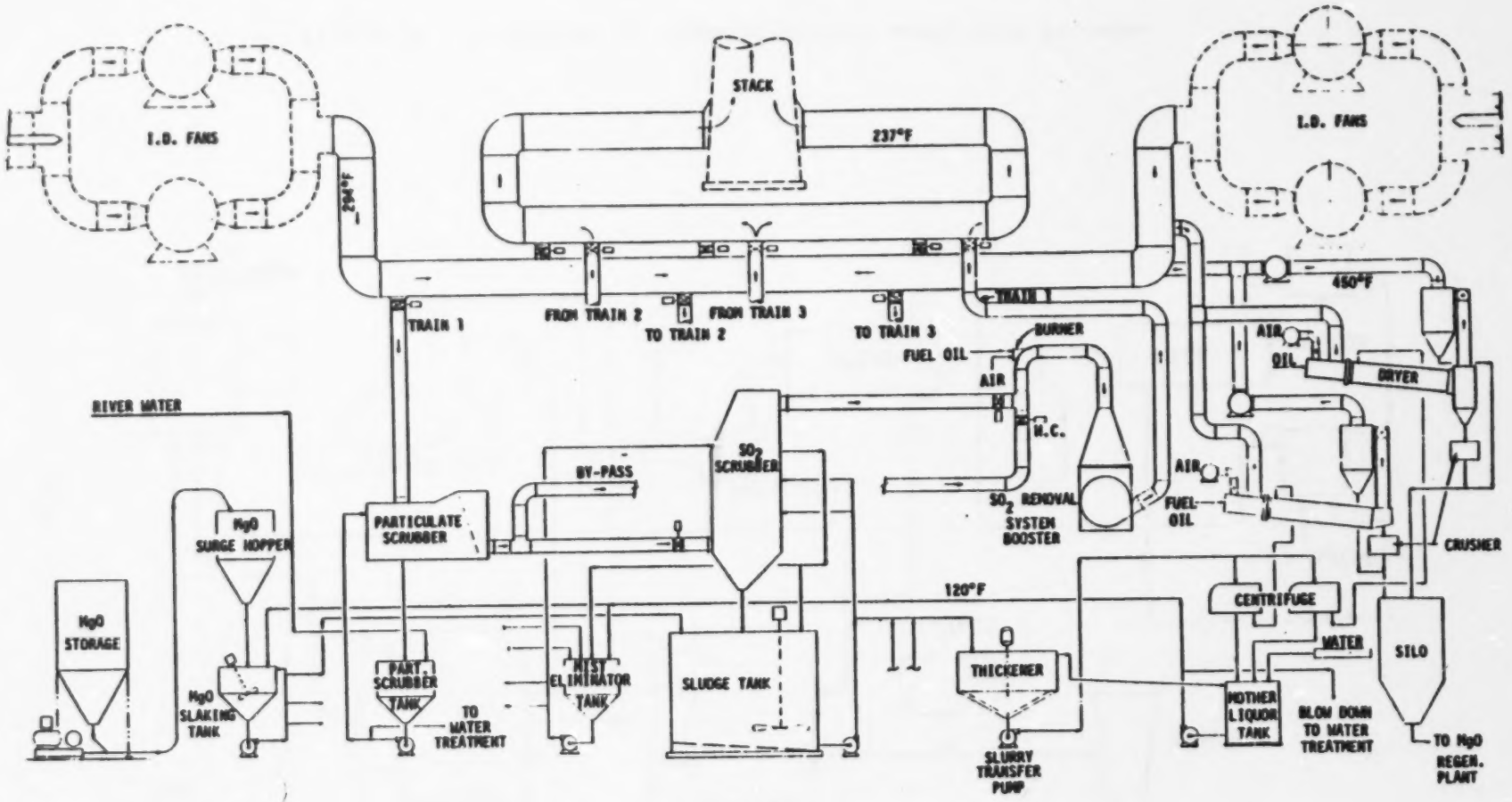

Figure D. Eddystone Generating Station.

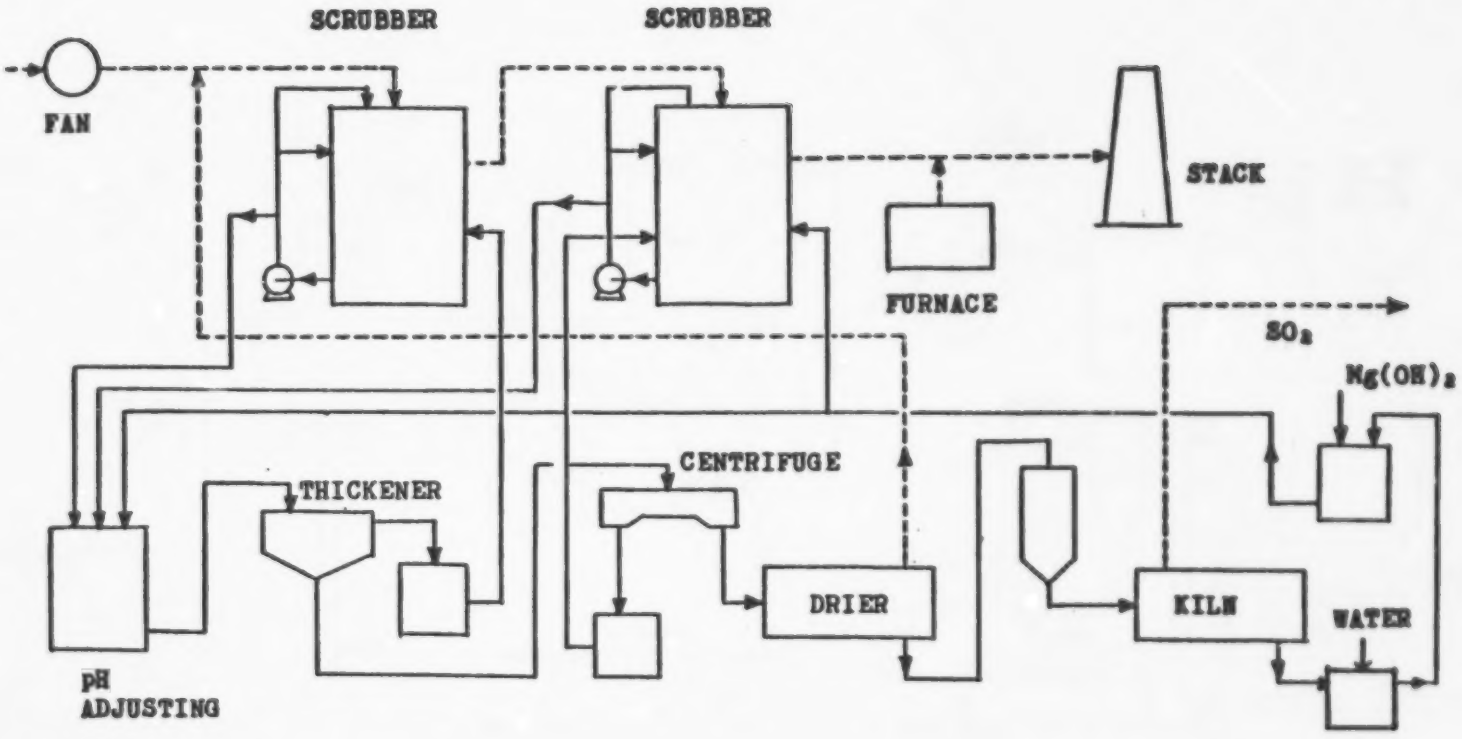

Figure E. Flowsheet of Chemico-Mitsui magnesium process.

APPENDIX B

APPENDIX B

Availability Factor Computation

1. Divide boiler capacity by the number of modules and obtain MW/module = x

2. Multiply boiler capacity by number of hours during selected period (MW) = a

3. Add all down times due to module trouble for all modules during selected period = b

4. Add all down times due to boiler trouble or reduction in electricity demand for all modules during period = c

5. Availability factor = $\frac{[a - x\ (b + c)]\ 100}{a - x\ c}$ = %

APPENDIX C

APPENDIX C. CONVERSION FACTORS USED IN THIS STUDY

Multiply	By	To Obtain
ft	x 0.30480	m
ft^3	x 28.317	liter
Btu/lb	x 0.55554	kcal/kg
	x 2.32593	kJ/kg
Btu/h	x $0.29300 \cdot 10^{-3}$	kW
acfm	x 1.6992	m^3/h
$lb/10^6$ Btu	x 1.80005	mg/kcal
	x 0.42993	mg/kJ
gal/min	x 3.78541	liter/min
	x 0.22712	m^3/h
gal/1000 ft^3	x 0.13368	liter/m^3
ft^3/min	x 1.6992	m^3/h
kcal	x 4.1868	kJ
kcal	x $1.163 \cdot 10^{-3}$	kWh
scfm	x 1.60776	Nm^3/h
gr/ft^3	x 2.288	g/m^3
gr/scfm	x 2.4182	g/Nm^3
in. H_2O	x 0.2488	kPa
in. H_2O	x 2.54	mbar
gr	x 0.0648	g
ton (short)	x 1.9072	metric ton

1 MW corresponds to 3200 to 3500 m^3/h flue gas (coal)

1 scfm = 28.317 l (60°F / 14.7 psia)
(15.5°C / 760.28 torr)

26.796 l (0°C/760 torr) = 0.026796 Nm^3

1 SO_2 = 2.927 g/liter

MW = electric power generated unless other wise specified